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PROCESS AND APPARATUS FOR C₂ RECOVERY

BACKGROUND OF THE INVENTION

1. Field of the Invention

This invention relates to an improved process for separating a hydrocarbon-bearing feed gas which contains methane and lighter components, (not necessarily all hydrocarbon components), C₂ (ethylene and ethane), and heavier hydrocarbon components into two fractions. The first fraction contains predominantly methane and lighter components and the second fraction contains the recovered desirable C₂ and heavier components. More particularly, this invention relates to a process and apparatus wherein the yield of C₂s is increased or alternatively energy consumption is reduced for a given C₂ recovery.

2. The Prior Art

Hydrocarbon-bearing gas may contain lighter components (e.g., hydrogen, nitrogen, etc.) methane, ethane, and/or ethylene, and a substantial quantity of hydrocarbons of higher molecular weight, for example, propane, butane, pentane and often their unsaturated analogs. Recent changes in ethylene/ethane demand have created increased markets for ethylene/ethane and have created a need for more efficient processes which yield higher recovery levels of this product. In more recent times, the use of cryogenic processes utilizing the principle of gas expansion through a mechanical device to produce power while simultaneously extracting heat from the system have been employed. The use of such equipment depends upon the pressure of the gas source, the composition of the gas and the desired end results. In the typical cryogenic expansion-type recovery processes used in the prior art, a gas stream under pressure is cooled by heat exchange with other streams of the process and/or external sources of cooling are employed such as refrigeration systems. As the gas is cooled, liquids are condensed and are collected and separated so as to thereby obtain desired hydrocarbons. The high pressure liquid feed is typically transferred to a demethanizer column after the pressure is adjusted to the operating pressure of the demethanizer. In

column after the pressure is adjusted to the operating pressure of the demethanizer. In such fractionating column the liquid feed is fractionated to separate the residual methane and lighter components from the desired products of ethylene/ethane and heavier hydrocarbon components. In the ideal operation of such separation processes, the vapors leaving the process contain substantially all of the methane and lighter components found in the feed gas and substantially no ethylene/ethane or heavier hydrocarbon components remain. The bottom fraction leaving the demethanizer typically contains substantially all of the ethylene/ethane and heavier hydrocarbon components with very little methane or lighter components which is discharged in the fluid gas outlet from the demethanizer.

A patentability search was conducted on the present invention and the following references were uncovered.

Inventor	Patent No.	Issue Date
Harandi	4,664,784	5/12/1987
Buck et al	4,895,584	1/23/1990
Campbell et al	5,771,712	9/01/1998
Wilkinson et al	5,699,507	6/30/1998

U.S. Patent No. 4,664,784 – Issued 05/12/1987
M.N. Harandi to Mobil Oil Corporation

In a reference directed to fractionation of hydrocarbon mixtures, teachings are found on column 4, line 32 et sequitur re: a zone (81) wherein a descending liquid heavy-ends portion contacts an ascending vaporous light-ends portion so as "...to aid in heat transfer between vapor and liquid." (column 4, line 44).

U.S. Patent No. 4,895,584 – Issued 01/23/1990
L.L. Buck et al to Pro-Quip Corporation

A reference that claims an improved process for hydrocarbon separation and teaches supplying of the liquids recovered from the light-ends fractionating column to the heavy ends fractionating column and directing part of the (C₂ containing) liquid from a

first step into intimate contact with a second residue, which liquid provides additional liquefied methane which acts with the partially condensed second residue as a direct contact refrigerant to thereby condense C₂ and heavier comprising hydrocarbons while methane itself is evaporated in the light-ends fractionation column.

On column 1, lines 56 – 67 the following teachings are found: "...feed gas is first cooled and partially condensed and delivered to a separator to provide a first residue vapor and a liquid containing C₂...Part of the liquid containing C₂ from the separator may be directed into a heavy-ends fractionation column wherein the liquid is separated into a second residue containing lighter hydrocarbons and C₂ containing products. A part of the first residue vapors with at least part of the partially condensed second residue are counter currently contacted and commingled in a light-ends fractionation column (emphasis added)..."

On column 2, lines 1-10 the following teachings are found: "The liquids recovered from the light-ends fractionation column are then fed to the heavy-ends fractionation column as a liquid feed. A portion of the liquids containing C₂ from the separator is fed into intimate contact with the second residue prior to discharging the commingled liquids and gases into the light-ends fractionation column to thereby achieve mass and heat transfer (emphasis added) to thereby liquefy a higher percent of the C₂ and heavier hydrocarbon components while the methane is vaporized" (column 2, lines 1-10).

The following Elcor Corporation references describe the recovery of C₃ and heavier hydrocarbons via processes wherein counter-current contact of a stream drawn from a deethanizer with a stream in a separator/absorber takes place:

U.S. Patent No. 5,799,507 – Issued 09/01/1998
J.D. Wilkinson et al to Elcor Corporation

See column 4, line 2 re: "...liquid portion of expanded stream commingles with liquids falling downward from the absorbing section..." I.o.w., the stream (36) from the

deethanizer (17) flows through heat exchanger (20) to become stream (36a) which flows into the upper section of separator (15) where it "...contacts the vapors rising upward through the absorption section" (column 5, lines 3-4).

U.S. Patent No. 5,771,712 – Issued 06/30/1998

R.E. Campbell et al to Elcor Corporation

This reference teaches essentially the same as Wilkinson et al.

None of the foregoing patents discussed above embody the present invention.

SUMMARY OF THE INVENTION

The present invention provides processes for increasing the ethylene and ethane component of the discharge from the process unit at reduced energy consumption than the prior art. The foregoing advantage is achieved in the present invention by a process in which the feed gas is first cooled and partially condensed and delivered to a separator to provide a first residue vapor and a first liquid containing C₂ which liquid also contains lighter hydrocarbons. A first part of the first liquid containing C₂ from the separator may be directed into a heavy-ends fractionation column, wherein the liquid is separated into a second residue containing lighter hydrocarbons and a second liquid product containing C₂. A second part of the first liquid from the separator is cooled. The second residue is cooled and partially condensed and then combined with the cooled second part of the first liquid providing, upon separation, a third residue and a third liquid. A first part of the third liquid is cooled and fed to the light-ends fractionation column. A second part of the third liquid is fed directly to the heavy-ends fractionation column. A part of the first residue vapor with a cooled first part of the third liquid are counter-currently contacted and commingled in a light-ends fractionation column to thereby provide fourth residue vapor and liquid which are separately discharged. Cooling the first part of the third liquid prior to its introduction into the light-ends fractionation column aids in mass and heat transfer. This cooling thereby provides for greater liquefaction of a higher percent of the C₂ and heavier hydrocarbon components

while the methane contained in the first part of the third liquid is vaporized. The fourth liquid recovered from the light-ends fractionation column is heated then introduced to the heavy-ends fractionation column as a feed.

A better understanding of the invention will be had with reference to the following description and claims, taken in conjunction with the attached drawings.

DESCRIPTION OF THE DRAWINGS

FIG. 1 is a schematic flow diagram illustrating a method of practicing a preferred embodiment of the invention.

FIG. 2 is a schematic flow diagram illustrating a variation in the preferred embodiment of the present invention.

DESCRIPTION OF THE PREFERRED EMBODIMENTS

The improved processes of the present disclosure include the steps of cooling a gaseous hydrocarbon-containing feed stream to form a first vapor stream and a first liquid stream. A first part of the first liquid stream is transferred to a heavy-ends fractionation column while the first vapor stream is transferred to the bottom of a light-ends fractionation column. The heavy-ends fractionation column overhead vapor, which consists mainly of methane, ethylene, and/or ethane, is cooled and partially condensed. The cooled heavy-ends fractionation column overhead is combined with a cooled second part of the first liquid stream. The resulting stream is fed to a separator and separated into a third residue vapor and a third liquid. A first part of the third liquid is cooled and fed to the upper portion of the light-ends fractionation column. The liquid flows downwardly within the light-ends fractionation column and contacts gaseous ethylene and/or ethane and heavier hydrocarbons that flow upwardly. The methane portion of the liquid stream is vaporized by absorbing heat from the gaseous ethylene/ethane and heavier hydrocarbons which causes the ethylene/ethane and heavier hydrocarbons to condense and exit at the bottom of the light-ends fractionation

column. The gaseous methane and lighter components within the light-ends fractionation column are removed from the overhead as a product of the process. The second part of the third liquid may be used to reflux the heavy-ends fractionation column. The fourth liquid at the bottom of the light-ends fractionation column is removed and used to cool other process streams; the thus-heated fourth liquid is fed to the upper portion of the heavy-ends fractionation column. The liquid at the bottom of the heavy-ends fractionation column is removed as a product of the process.

The improved process of this invention is illustrated in a first embodiment in FIG. 1. The incoming gas stream 1 at a temperature of 120°F and a pressure of 827 psia passes through heat exchanger 38, so that the temperature thereof is reduced to about -72°F with attendant partial condensation. Pressure is reduced as the gas flows through the heat exchangers resulting in a pressure of 812 psia at -72°F at which the raw gas is delivered into a separator 44. Within separator 44 the cooled gas stream is separated into a first liquid stream (stream 4) and a first residue vapor, stream 3. Stream 3 is passed through a turbo expander 46. The shaft of turbo expander 46 is connected directly to the shaft of the booster compressor 32. From the turbo expander, the first residue gas having a temperature of about -163°F at 200 psia passes by way of stream 5 into a light-ends fractionation column 52.

From separator 44 a first part of the first liquid containing C₂ is conducted into a heavy-ends fractionation column 56 by way of stream 4A. A second part of the first liquid containing C₂ from stream 4 is channeled by way of stream 4B through heat exchanger 42 where its temperature is decreased. The cooled liquid exits the heat exchanger and combines with the cooled residue stream 14 to form stream 16.

The second residue from heavy-ends fractionation column 56, having a temperature of about -132°F, is fed by way of stream 14 through heat exchanger 42, combines with the remainder of the liquid containing C₂ from stream 4B above, and by

way of stream **16** into the reflux separator **57**. A first part of the third liquid from the reflux separator **57** is routed by stream **23** through heat exchanger **42** where its temperature is reduced. This liquid stream is then passed as stream **23A** into the light-ends fractionation column **52**. The liquid from stream **23A** passes downwardly through the light-ends fractionation column **52** and encounters the rising first residue gas from stream **5** so that mass and latent heat transfer occur. The second part of the third liquid from the reflux separator **57** is routed by stream **26** to the heavy-ends fractionation column **56**.

The light-ends fractionation column **52** functions as a combination heat and mass transfer device. The column has two feed streams; that is, streams **5** and **23A**, and two product streams; that is, streams **10** and **9**. The light-ends fractionation column **52** consists of at least one, and preferably more, theoretical liquid-vapor equilibrium stages.

Vapor enters the light-ends fractionation column by way of stream **5** as a bottom feed while the top feed is by way of stream **23A** which is a liquid enriched by condensed methane. The methane and lighter constituents and un-recovered ethylene and ethane, exit as a dew point vapor as a fourth residue (stream **9**) from the top tray or separation stage of the light-ends fractionation column **52**.

The top feed through stream **23A** into the light-ends fractionation column **52** and particularly the methane content thereof serves as a reflux in the column. In flowing from stage to stage within column **52**, the liquid methane is vaporized and in turn the liquid is progressively enriched in ethylene and ethane condensed from the upflowing bottom feed vapor from stream **5**.

The fourth liquid stream from the light-ends fractionation column **52**, stream **10**, provides process cooling in exchanger **42** while it is itself warmed and then fed to the heavy-ends fractionation column **56** for further separation.

The fourth residue gas (stream 9) discharged from light-ends fractionation column 52 passes through exchangers 42 and 38 and exits the heat exchanger system as stream 19. The third residue gas vapor in stream 18 exiting the reflux separator 57 also pass through exchangers 42 and 38 and exit the heat exchanger system as stream 28. The warmed vapor from the light-ends fractionation column (stream 19) is compressed in compressor 48 to the same pressure as stream 28 and combined with stream 28 to form stream 30. The combined vapors of stream 30 are compressed in the booster compressor 32. At this stage, the methane rich off-gas in stream 21 has a temperature of 103°F and a pressure of 187 psia. If it is desired to return the discharge gas to the same system from which the raw gas was taken, such as for further transportation of the gas, the pressure will need to be raised back to that substantially equal to the incoming pressure of 827 psia in stream 1.

The second liquid discharge, rich in C₂ content, from the lower end of the heavy-ends fractionation column 56 is passed by way of stream 15 and exchanger 38 to product discharge stream 22.

The result of a simulation of the process of FIG. 1 is set forth in Table 1A wherein the moles per hour of various constituents of the streams are set forth. The process achieves a recovery of about 97.37 percent of the C₂ content of the feed gas in addition to substantially complete recovery of the C₃ and heavier hydrocarbon components of the feed gas stream into the less volatile fraction (product).

Table 1B relates the moles per hour of various constituents of the stream of the process of FIG. 1 when the process of FIG. 1 is applied to a feed gas stream that is enriched in ethane and heavier components.

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FIG. 2 shows an alternate embodiment of the invention. The components of the process of FIG. 2 having the same basic structure and function of those of the system of FIG. 1 are given like numbers. The process is as described with reference to FIG. 1, except that the booster compressor **32** is placed on the feed gas (stream **1**) and streams **9** and **18** are combined prior to exchanger **42**.

Table 2, shows the result of a simulation of the system of FIG. 2. Table 2 provides the moles per hour of various constituents for the various streams of this embodiment of the process. The process achieves a recovery of about 91.64 percent of the ethylene and 96.77 percent of the ethane content of the feed gas in addition to substantially complete recovery of the C₃ and heavier hydrocarbon components of the feed gas stream in to the less volatile fraction (product).

The process has been illustrated using various standard components employed for the sequence of treating steps with it being understood that the process may be practiced utilizing different physical apparatus. For instance, the turbo expander can, in many instances, be eliminated or replaced by a Joule-Thomson isenthalpic control valve. The difference is that where the expander is eliminated or where the Joule-Thomson valve is substituted for the turbo expander, normally greater inlet and refrigeration compression duties are required.

A different arrangement has been shown in the alternate embodiment for cooling the second residue effluent and thus providing reflux to the light-ends fractionation and heavy-ends fractionation columns.

Some of the processes in each instance may use multiple turbo expanders. The desirability of the use of multiple turbo expanders is predicated primarily upon the amount of hydrogen content of the inlet gas in stream **1**. It is understood that, according to the inlet gas content, only single turbo expanders may be employed in practicing the

process; or, in some instances as previously indicated, turbo expanders may be eliminated completely or substituted by one or more Joule-Thomson isenthalpic expansion valves.

An important feature of the process is the employment of the light-ends fractionation column **52** which functions as a combination heat and mass transfer device. The use of the reflux in the top stage means that the liquid methane of the reflux is vaporized; and in turn the liquid is progressively enriched in ethylene and ethane condensed from the upflowing bottom feed vapor to thereby recover a higher percent of the C₂ components.

While the invention has been described with a certain degree of particularity, it is manifest that many changes may be made in the details of construction and the arrangement of components without departing from the spirit and scope of this disclosure. It is understood that the invention is not limited to the embodiments set forth herein for purposes of exemplification, but is to be limited only by the scope of the attached claim or claims, including the full range of equivalency to which each element thereof is entitled.

TABLE 1A

THE PRO-QUIP CORPORATION										
STREAM	STREAM NUMBER									
NAME	1	3	4	5	9	10	14	16	18	23
NITROGEN	99 17	93 86	5 31	93 86	92 74	2 50	7 81	7 81	6.43	1.38
CARBON DIOXIDE	8.64	6 19	2 45	6 19	1 89	5 83	1 95	1 95	0.42	1.53
METHANE	7552.91	6526.79	1026 12	6526 79	6374 89	1115 68	2131 34	2131 34	1168 58	963 78
ETHANE	486 41	272 65	213 76	272 65	9 22	299 03	39 18	39 18	3 57	35 60
PROPANE	198 31	56.60	141 71	56 60	0 04	58 45	1 92	1 92	0 03	1 89
I-BUTANE	36.66	5 59	31.07	5.59	0 00	5.67	0 08	0 08	0 00	0.08
N-BUTANE	63 30	7.19	56 11	7 19	0 00	7 27	0 08	0 08	0 00	0.08
I-PENTANE	20 83	1.16	19 67	1 16	0.00	1 17	0 01	0 01	0 00	0.01
N-PENTANE	20.63	0.86	19 77	0 86	0.00	0 86	0 00	0 00	0 00	0 00
HEXANE	19.29	0 29	19 00	0 29	0 00	0 29	0 00	0 00	0 00	0 00
TOTAL LBMOL/HR	8525 10	6971 27	1553 83	6971 27	6478 78	1496.83	2,182 36	2,182 36	1179.02	1004.35
MASS FLOW LB/HR	160249	119227	41022	119227	105232	30726	35770	35770	19055	16732
VOLUME FLOW MMSCFD	78	63	-	-	59	-	20	-	11	-
MOL MOLE, WT	18 80	17 10	26 40	17 10	16 24	20 53	16 39	16 39	16 16	16.66
DENSITY LB/FT ³	2 83	5 92	26 30	1.51	1.32	26 94	2 06	4 01	2 35	20 51
TEMPERATURE °F	120	-72	-72	-163	-178	-165	-132	-153	-153	-153
PRESSURE PSIA	827.00	812 00	812 00	200 00	193 00	385 00	330 00	328 00	328 00	353 00

STREAM	STREAM NUMBER					Percent Recovered to Volatile Fraction	Percent Recovered To Less Volatile Fraction
NAME	25	26	15	22	21		
NITROGEN	0 00	0 00	0 00	0 00	99 17	100 00%	0 00%
CARBON DIOXIDE	0 00	0 00	6 33	6 33	2 31	26 71%	73 30%
METHANE	0 00	0.00	10 46	10 46	7543 46	99 87%	0 14%
ETHANE	0.00	0 00	473 61	473.61	12 79	2 63%	97 37%
PROPANE	0.00	0 00	198 24	198 24	0 07	0 04%	99 96%
I-BUTANE	0.00	0 00	36 66	36.66	0 00	0 00%	100 00%
N-BUTANE	0 00	0 00	63 30	63 30	0 00	0 00%	100 00%
I-PENTANE	0 00	0 00	20 83	20 83	0 00	0 00%	100 00%
N-PENTANE	0 00	0 00	20 63	20 63	0 00	0 00%	100 00%
HEXANE	0 00	0 00	19 29	19 29	0 00	0 00%	100 00%
TOTAL LBMOL/HR	0 00	0 00	868.31	868 31	7657 80		
MASS FLOW LB/HR	0	0	35978	35978	124286		
VOLUME FLOW MMSCFD	-	-	-	-	70		
MOLE WT	16 66	16 66	41.44	41 44	16.23		
DENSITY LB/FT ³	20.51	20 51	30 11	27.97	1.03		
TEMPERATURE °F	-153	-153	71	100	167		
PRESSURE PSIA	353 00	353 00	500 00	495 00	413 41		

TABLE 1B

THE PRO-QUIP CORPORATION

STREAM	STREAM NUMBER									
	1	3	4	5	9	10	14	16	18	23
NITROGEN	345 88	280 87	65 02	280 87	275.89	16.91	19.90	84.91	69.99	11.93
CARBON DIOXIDE	327 77	161 06	166 70	161.06	48 44	254.24	37.16	203 85	26 83	141.62
METHANE	24864 18	16379 96	8484 21	16379 96	17115 17	6530 39	8271 66	16754 42	7672.43	7265 60
ETHANE	3696 03	1309.26	2386 76	1309.26	179 47	3053 40	131 47	2518 25	113 74	1923.61
PROPANE	2012 72	363 60	1649 12	363 60	10 96	1673 25	11 09	1660 25	9 49	1320 61
I-BUTANE	385.41	40 55	344.87	40 55	0 43	316 09	0 54	345 41	0 46	275.96
N-BUTANE	612.71	50 73	561 98	50 73	035	500 03	0 49	562 47	0 41	449.65
I-PENTANE	151.53	7 05	144 48	7 05	0 02	122 62	0 03	144 51	0.03	115.59
N-PENTANE	115 29	4.29	111 00	4 29	0 01	93 09	0 01	111 02	0 01	88.80
HEXANE	98 82	1 67	97 15	1.67	0.00	79 39	0.00	97.15	0.00	77 72
HYDROGEN SULFIDE	0.00	0 00	0 00	0 00	0.00	0 00	0 00	0 00	0 00	0.00
CARBONYL SULFIDE	3 29	0 68	2 61	0 68	0 03	2 74	0 03	2 64	0 03	2 09
TOTAL LBMOL/HR	32613.64	18599.74	14013 90	18599 74	17630 77	12642 15	8472 40	22484 88	7893 41	11673 17
MASS FLOW LB/HR	708883	339451	369432	339451	290366	352036	139402	508812	130123	302951
VOLUME FLOW MMSCFD	297	169	-	-	161	-	77	-	72	-
MOLE. WT.	21 74	18.25	26.36	18 25	16.47	27.85	16 45	22 63	16.48	25.95
DENSITY LB/FT3	4 20	6 88	24.02	1.76	1.46	32.32	2 07	6 70	2.06	30.02
TEMP °F	120	-40	-40	-133	-149	-138	-129	-131	-131	-131
PRESSURE PSIA	978 00	966 35	966 35	242 00	237 00	375 00	335 00	330 00	330 00	370 00

STREAM	STREAM NUMBER					Percent Recovered to Volatile Fraction	Percent Recovered To Less Volatile Fraction
	25	26	15	22	21		
NITROGEN	2 98	2 98	0 00	0 00	345 88	100 00%	0 00%
CARBON DIOXIDE	35 40	35 40	252 48	252 48	75 27	22 97%	77 03%
METHANE	1816 40	1816 40	75 13	75 13	24787 60	99 69%	0 30%
ETHANE	480 90	480 90	3402 83	3402 83	293 21	7 93%	92 07%
PROPANE	330 15	330 15	1992 30	1992 30	20 46	1 02%	98 99%
I-BUTANE	68 99	68 99	384 53	384 53	0 89	0 23%	99 77%
N-BUTANE	112 41	112 41	611 95	611 95	0 76	0 12%	99 88%
I-PENTANE	28 90	28 90	151 49	151 49	0 04	0 03%	99 97%
N-PENTANE	22 20	22 20	115 28	115 28	0 02	0 02%	99 98%
HEXANE	19 43	19 43	98 82	98 82	0 00	0 00%	00 00%
HYDROGEN SULFIDE	0 00	0 00	0 00	0 00	0 00	0 00%	00 00%
CARBONYL SULFIDE	0 52	0 52	3 23	3 23	0 06	1 84%	98 16%
TOTAL LBMOL/HR	2918 29	2918 29	7088 04	7088 04	25524 18		
MASS FLOW LB/HR	75738	75738	288372	288372	420489		
VOLUME FLOW MMSCFD	-	-	-	-	232		
MOLE WT	25 95	25 95	40 68	40 68	16 47		
DENSITY LB/FT3	30 02	30 02	30 09	27 83	0 88		
TEMPERATURE °F	-131	-131	72	100	115		
PRESSURE PSIA	370 00	370 00	500 00	490 00	317 14		

TABLE 2

THE PRO-QUIP CORPORATION

STREAM	STREAM NUMBER									
NAME	1	3	4	5	9	10	14	16	18	23
HYDROGEN	1274.20	1203.16	71.03	1203.16	1200.28	3.85	29.12	75.29	73.92	0.96
NITROGEN	197.10	165.03	32.07	165.03	162.81	5.39	17.96	38.80	34.30	3.16
CARBON MONOXIDE	13.01	10.54	2.47	10.54	10.36	0.52	1.53	3.13	2.65	0.34
METHANE	3194.56	1790.74	1403.81	1790.74	1992.70	641.30	1485.42	2397.90	1197.69	843.29
ETHYLENE	672.81	127.55	545.26	127.55	29.42	356.01	39.41	393.82	26.82	257.87
ETHANE	1402.52	155.95	1246.57	155.95	21.51	711.58	34.92	845.19	23.80	577.13
PROPENE	195.47	5.89	189.58	5.89	0.24	92.39	0.64	123.86	0.41	86.74
PROPANE	156.55	3.57	152.98	3.57	0.12	73.40	0.35	99.79	0.22	69.96
I-BUTANE	1.51	0.01	1.50	0.01	0.00	0.70	0.00	0.98	0.00	0.68
N-BUTANE	81.73	0.45	81.28	0.45	0.00	37.57	0.02	52.86	0.01	37.13
N-PENTANE	28.36	0.03	28.33	0.03	0.00	12.97	0.00	18.42	0.00	12.94
TOTAL LBMOL/HR	7217.81	3462.92	3754.89	3462.92	3417.45	1935.67	1609.35	4050.03	1359.84	1890.19
MASS FLOW LB/HR	142766	44774	97992	44774	40727	52131	26634	90329	21995	48083
VOLUME FLOW MMSCFD	66	32	-	-	31	-	15	-	12	-
MOLE WT	19.78	12.93	26.10	12.93	11.92	26.93	16.55	22.30	16.10	25.44
DENSITY LB/FT ³	2.13	2.82	28.62	0.71	0.63	33.86	1.07	3.84	1.03	31.33
TEMPERATURE ° F	100	-89	-89	-171	-183	-175	-146	-152	-152	-152
PRESSURE PSIA	581.00	726.00	726.00	148.60	145.00	213.00	185.00	181.00	181.00	213.00

STREAM	STREAM NUMBER					Percent Recovered	Percent Recovered
NAME	25	26	15	22	21	to Volatile Fraction	To Less Volatile Fraction
HYDROGEN	0.41	0.41	0.00	0.00	1274.20	100.00%	0.00%
NITROGEN	1.34	1.34	0.00	0.00	197.11	100.00%	0.00%
CARBON MONOXIDE	0.14	0.14	0.00	0.00	13.01	100.00%	0.00%
METHANE	356.91	356.90	4.21	4.21	3190.39	99.87%	0.13%
ETHYLENE	109.14	109.14	616.58	616.58	56.24	8.36%	91.64%
ETHANE	244.26	244.27	1357.22	1357.22	45.32	3.23%	96.77%
PROPENE	36.71	36.71	194.82	194.82	0.65	0.33%	99.67%
PROPANE	29.61	29.61	156.21	156.21	0.34	0.22%	99.78%
I-BUTANE	0.29	0.29	1.51	1.51	0.00	0.00%	99.97%
N-BUTANE	15.71	15.71	81.71	81.71	0.02	0.02%	99.98%
N-PENTANE	5.48	5.48	28.36	28.36	0.00	0.00%	100.00%
TOTAL LBMOL/HR	800.00	800.00	2440.62	2440.62	4777.28		
MASS FLOW LB/HR	20351	20351	80146	80146	62622		
VOLUME FLOW MMSCFD	-	-	-	-	-		
MOLE WT	25.44	25.44	32.84	32.84			
DENSITY LB/FT ³	31.33	31.33	30.51	23.64			
TEMPERATURE ° F	-152	-152	-7	74			
PRESSURE PSIA	213.00	213.00	585.00	580.00			